Behaviour of Three Phase Fluidized Bed with Regular Particles

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By

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CERTIFICATE

This is to certify that the work in this thesis entitled "Behaviour of Three Phase Fluidized Bed with Regular Particles" submitted by Sasmita Hembram (108CH020) in partial fulfillment of the requirements of the prescribed curriculum for Bachelor of Technology in Chemical Engineering Session 2008-2012 in the department of Chemical Engineering, National Institute of Technology,Rourkela is an authentic work carried out by her under my supervision and guidance. To the best of my knowledge the matter embodied in the thesis is her bona fide work.

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ABSTRACT

Fluidized beds have got a great application in the industries. So it is very important to understand aspects like type of distributor, particle shape and size, type of fluid etc. for the proper functioning of the fluidized beds. In the present work, hydrodynamics of the fluidized bed with regular particles was studied. The hydrodynamic behaviour like pressure drop, minimum liquid fluidization velocity and bed expansion ratio of a co-current three-phase fluidized bed have been determined.

Experiments were carried out using raschig ring of 3.3 mm and 5.9 mm as solid particles, water as liquid and compressed air as gas .Variation of pressure drop with liquid velocity at constant gas velocity and variation of bed expansion ratio with liquid velocity at constant gas velocity and minimum liquid fluidization velocity with gas velocity was investigated. It has been observed that the minimum liquid fluidization velocity decreases with the increase in superficial gas velocity. The mass of the bed particles does not affect the minimum liquid fluidization velocity. The results obtained have been compared with other correlations and have been found to agree well.

Keywords : Fluidization, Hydrodynamics, Pressure drop, Minimum liquid fluidization velocity, Bed expansion, Regular particles.

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NOMENCLATURE

ΔP -Bed Pressure Drop, kPa

β- Bed Expansion Ratio

H –Fluidized Bed Height, cm

Hs- Static Bed Height, cm

ε_g- Gas holdup

ε_l- Liquid Holdup

ε_s- Solid Holdup

ε- Porosity

ρ- Density, Kg/m^3

μ- Viscosity, Pa.s

Ul- Liquid velocity, m/s

Ug- Gas velocity, m/s

U_{lmf}- Minimum liquid fluidization velocity, m/s

dp- Diameter of particle, mm

CHAPTER-1

INTRODUCTION

Fluidization is a process in which solids behave like fluid due to the passing of fluids through it. In this technique fluid is passed through a bed of solids when after a certain velocity called minimum fluidizing velocity the bed of particles rise from its stationary condition and gets fluidized. It has got application in many fields and has been an important topic for research. In spite of lot of research activities carried out for the understanding of fluidization technology in the past few decades, several aspects relating to the effect of distributor, irregular and regular shape particles as bed material, liquid viscosity and surface tension, scaling up of a developed system for industrial application are not fully investigated (Jena et al.,2009) This technique of contacting has its application in fields like Waste water Treatment, Coating of materials, Fischer-Tropsch process, Coal liquefaction (Jena et al.,2009) .

1.1 Types of fluidization

Fluidization is mainly of two types i.e. Particulate fluidization and aggregative fluidization. Particulate fluidization occurs when the solid and fluid density difference is not much and the solids are smaller in size. In this fluidization the fluid velocity required to fluidize the bed is not much (example- liquid-solid systems).

Aggregative fluidization occurs when the solid and fluid density difference is more and solids are larger in size. In this fluidization the fluid velocity required to fluidize the bed is quite high (example- gas-solid systems). In this case when fluidization occurs then bubbles form in between the solids because of the large particle size and high liquid velocity. These bubbles carry little or no solid particles with them. When these bubbles rise from the bed then they eventually break at the surface of bed. The superficial fluid velocity in which fluid bubbles form is called the minimum bubbling velocity. Generally a bubbling fluidized bed is considered to be undesirable for industrial application. (Narayanan et.al, 2009)

1.2 Advantages of Fluidized Bed

Three-Phase Fluidized bed has proved to be better than conventional reactors and has got many advantages like.

- It can maintain uniform temperature.
- Bed plugging and channelling can be reduced due to the movement of solids.
- Pressure drop is less due to which pumping cost can be reduced etc.
- High turbulence.
- Flexibility of better mixing.
- It can use fine catalyst particles, which minimizes the intraparticle diffusion.

However there are some disadvantages to fluidized bed like entrainment and carryover of particles, due to particle motion attrition of catalyst can occur.

1.3 Three Phase Fluidization

Three phase fluidization also known as gas-liquid-solid fluidization. In Three Phase Fluidization bed of solid particles are made to suspend in liquid and gas media due to the net force of the fluid on solids. The solids are supported by a plate which is porous in nature called distributor through which liquid and/or gas is passed. The bed of solids will remain in their position if the velocity of the fluid is not enough. The force of the fluid should be able to counter-balance the weight of the solids then only the solids will get fluidized. As the fluid velocity is increased the pressure or the force of the fluid which is forced against the gravitational force will increase. Due to the increase in velocity of the fluid a stage will come when the bed will begin to expand and after which contents will move like an agitated tank. At the velocity called minimum fluidization velocity the weight of the bed will be equal to the pressure of the fluid. Three phase reactor can be classified into two types of reactor depending on its density and volume fraction of particles. They are fluidized bed reactors and slurry bubble column reactors. In fluidized bed reactors the density of the solids are much more than the fluid where the volume fraction of solids lies between 0.2-0.6 (Pandey et al.,2010) and the particle size is in above 150μm. While in slurry bubble column reactor the density of the solids are slightly more than the fluid ,volume fraction of the particles ,volume fraction of solids is less than 0.15 and particle size lies in the range of 5-150μm.

1.3.1 Modes of Three Phase Fluidization

In Three Phase Fluidization all contacting measures need to be checked. Fluidized beds are categorized by the direction of flow: co-current up-flow, co-current down-flow, counter current, and liquid batch with gas up-flow. The present work is limited to co-current up-flow of the liquid and gas, the commonly used type of three-phase flow (Jena et al.,2009). In a catalytic reaction, there are eight key steps which needs to be considered for the scale-up and design of a fluidized bed reactor which are as follows:

• Mass transfer between gas-liquid interfaces

• Mass transfer between liquid-solid interfaces

• Diffusion from the catalyst surface

• At the surface of catalyst, adsorption of reactants takes place

• Reaction at the catalyst surface

• Desorption of adsorbed products

• Transfer of products from catalyst interior sites to its outer surface

• Transfer of products from catalyst surface to bulk fluid.

Each of these aspects needs to be properly studied so that we can clearly understand the nature of fluidized bed. The first of these steps is commonly the rate-controlling one, and it depends on the hydrodynamic behaviour of bed. So it is very important to know the bed hydrodynamics, mainly the bed expansion, pressure drop, the phase hold-ups and the minimum fluidization velocity. For a given reactor volume and fluid flow rates, a cloud of small, spherical, slow-moving bubbles will have a greater opportunity to allow mass transfer than a few large, quick moving slugs. (Jena et al.,2009) .

1.3.2 Flow Regime

Knowledge of flow regimes in fluidized bed is required to be known for its stable operation with different types of operating variables. Fan (1989) identified three flow regimes of three-phase fluidized beds: bubbling, slugging, and transport. Bubbling regime is again categorised into dispersed bubble and coalesced bubble regimes. The separation between regimes has not always been well defined. Zhang (1996) and Zhang et al. (1997) has seven different flow regimes for gas-liquid-solid co-current fluidized beds and has found different quantitative methods for determining the transitions as under:

• Dispersed bubble flow: It generally corresponds to low gas velocities and high liquid velocities. In dispersed bubble flow it has small bubbles of uniform size. Little bubble coalescence in spite of high bubble frequency.

• Discrete bubble flow: It corresponds to low liquid and gas velocities. It also relates to small bubble size with lower bubble frequency.

• Coalesced bubble flow: It generally relates to low liquid velocities and intermediate gas velocities. However the bubble size is big with increased bubble coalescence.

• Slug flow: Not much seen in industrial applications. In this regime it has large bullet shaped bubbles with a diameter of that of the column and length which exceeds the column diameter. In the wakes of the slugs some smaller bubbles can also be observed.

• Churn flow: It is quite similar to that of slug flow regime. As the gas flow is increased, an increase in liquid flow near the wall also increases in downward direction. Darton (1985) explained it as the transition between bubbling and slug flow on the basis of two-phase fluidized systems.

• Bridging flow: Its a transitional regime between the churn flow and the annular flow. In this flow liquid and solids form 'bridges" in the reactor which gets continuously broken and re-formed.

• Annular flow: At very high gas velocities, a continuous gas phase occurs in the core of the column.

The bubble sizes and shapes of bubbling, slugging, and transport regimes differ from each other. As earlier mentioned, slug flow has bullet shaped bubbles whose cross-sectional dimension equals to that of reactor column. In the transport regime, bubbles are not much formed since the gas forms a continuous phase as it conveys liquid droplets and solid particles through the fluidized bed. In the bubbling regime, under dispersed bubble flow, the bubbles tend to be spherical, small and relatively uniform in size. In coalesced flow the bubbles tend to grow larger with a wider size distribution. Spherical-cap or spheroid bubbles are also commonly found, and these can have significant wakes that also affect the reactor performance (Jena et al.,2009). Wakes are often responsible for increased particle mixing and is the reason for some beds to contract initially when the gas flow is increased (Jena et al.,2009).

Particle action is another important aspect of the flow regimes. In a transport system, many particles are entrained and gets removed from the system. Hence particles need to be replenished continuously. While in a bubbling bed, entrainment of particles is not much, that's why replenishing of particles is not much required. Although most classifications of the fluidization regime are related to liquid and gas superficial velocities but it is also important to consider particle size, shape, and density (Jena et al.,2009).

1.4 Terms related to Fluidization Phenomena

1. Minimum Fluidization Velocity (U_{mf}) – The minimum superficial velocity at which the bed just gets fluidized. At this velocity weight of the bed just gets counterbalanced by the pressure of the fluid.

2. Bed Pressure Drop (ΔP) – Measures the total weight of the bed in combination with the buoyancy and phase holdups.

3. Bed Expansion Ratio (β) – It the ratio of fluidized bed height to the initial bed height.

$$\beta = H/Hs \qquad \textbf{-(1)}$$

4. Gas Holdup (ε_g) – It measures the volume fraction of gas. It is the ratio of volume of gas to the total volume of bed.

5. Liquid Holdup (ε_l) – It measures the volume fraction of liquid. It is the ratio of volume fraction of liquid to the total bed volume.

6. Solid Holdup (ε_s) – It measures the volume fraction of solid. It is the ratio of volume fraction of solid to the total volume of bed.

$$\varepsilon s + \varepsilon l + \varepsilon g = 1 \qquad \textbf{-(2)}$$

7. Porosity (ε) – It is the volume occupied by both liquid and gas.

$$\varepsilon = \varepsilon l + \varepsilon g = 1 - \varepsilon s \qquad -(3)$$

1.5 Scope and objective of the present investigation

The aim of the present work has been précised below:

- Design a fluidized bed system with an air sparger and distributor arrangement which ensures less pressure drop across the column and uniform distribution of the fluids.
- Hydrodynamic study of gas-liquid-solid fluidized bed with regular particles- bed pressure drop, bed expansion and minimum fluidization velocity.
- Examining the effect of the gas velocity on hydrodynamic properties studied.

The present work is focussed on understanding the hydrodynamic behaviour in a three phase fluidized bed. Experiments were performed in a fluidized bed of height 1.4 m with diameter of 0.1 m. Trial experiments were performed with 2.18 mm glass beads to check proper functioning of the fluidized bed system. Raschig ring of size 3.3 mm and 5.9 mm are used as the solid phase. The fluidization operation has been carried out with co-current insertion of water as the continuous phase and air as the discrete phase. Superficial velocity of both liquid and gas has been varied in the range of 0-0.175 m/s and 0-0.127 m/s. The static bed heights of the solid phase in the fluidized bed used are taken as 15.4 cm, 15.6 cm, 21.4 cm, 26.4 cm , 31.4 cm

1.6 Thesis Layout

The second chapter provides a comprehensive literature survey related to the hydrodynamic characteristics. It includes the experimental as well as the computational aspect of gas-liquid-solid fluidization. The third chapter deals with the experimental setup and techniques used to study the hydrodynamic properties. Chapter four deals with the hydrodynamic study- bed pressure drop, minimum liquid fluidization velocity, bed expansion ratio. Chapter five infers the conclusion drawn from the result obtained from the study and also presents the scope of the future work.

CHAPTER-2

LITERATURE REVIEW

The first fluidized bed was first found by Winkler in 1921 and industrial fluidized bed was first used as large-scale in Winkler gasifier in 1926 (Kunii and Levenspiel, 1991). Fluidized bed catalytic cracking of crude oil to gasoline (FCC) was commercialized in 1942, and is still the major application of fine-powder fluidization. Several catalytic applications such as acrylonitrile synthesis, phthalic anhydride and Fischer-Tropsch synthesis of liquid fuels from coal-based gas extended the range following the FCC (Jena et al.,2009). Lurgi commercialized the circulating fluidized bed (CFB) in the 1970's, for coarse powders, which would operate above the terminal velocity of all the bed particles. Polyethylene was produced in a fluidized bed, and the technology is widely used in industry. Commercialization of circulating fluidized bed was done in 1980's for the combustion and production of polypropylene in fluidized beds. New areas of application in fluidization were production of semiconductors and ceramic materials by chemical vapour deposition and in biological applications the use of liquid fluidized beds (Jena H. M ,2009).

The successful design and working of a gas-liquid-solid fluidized bed system depends on its ability to accurately predict the fundamental characteristics of the system mainly the hydrodynamics, the mixing of individual phases, and the heat and mass transfer characteristics (Jena et al.,2009a).Three-phase fluidized beds are also often used in physical operations (Jena et al.,2008b).

In the case of fixed bed operation, both cocurrent and countercurrent gas-liquid flow are allowed, and for each of these both bubble flow, in which the liquid is the continuous phase and the gas dispersed, and trickle flow. In which the gas forms a continuous phase and the liquid is more or less dispersed (Epstein, 1981).

In a chemical process mass transfer is the rate-limiting step, so it is very important to be able to estimate the gas hold-up as this relates directly to the mass transfer.

CHAPTER-3

EXPERIMENTAL SET-UP

A fluidized bed consists of three sections, ie., the test section, the gas-liquid distributor section, and the gas-liquid disengagement section. Fig. 1 shows the schematic representation of the experimental setup of three-phase fluidized bed. Fig. 2 shows the photographic representation of the setup. Fluidization takes place in the middle column called the test section which is a vertical cylindrical Plexiglas column of 0.1 m diameter and 1.4 m height consisting of three pieces of persepex columns with a flange and nut bolt arrangement having rubber gasket in-between. The gas-liquid distributor is located at the bottom of the test section and is designed in such a way that uniform distribution of liquid and gas mixture enters into the test section. The distributor section is a fructo-conical of 0.31 m in height, having a divergence angle of 4.5°. The liquid inlet of internal diameter 0.0254 m is located at the lower cross-sectional end. The higher cross-sectional end is fitted to the test section, with a perforated distributor plate made of G. I sheet of thickness 0.001 m, and of diameter 0.012m . The size of the holes has been increased from the inner to the outer circle. This configuration of distributor plate keeps less pressure drop at the distributor plate and gives a uniform flow of the gas-liquid mixture into the test section. Usually the flow rate is higher at the centre than towards the wall thus driving the gas bubbles in the central zone to a greater extent causing there a higher holdup of gas. To avoid this unequal distribution at the entrance of the test section, the distributor plate has been designed in such a way that relatively uniform flow can be achieved throughout the cross-section. Figs. 3 and Fig .4 represent the photographic view of the gas-liquid distributor section and the distributor plate. An antenna-type air sparger of Fig. 5 of 0.09 m diameter with 50 number of 0.001 m holes has been fixed below the distributor plate with a few layers of plastic and glass beads in between for the generation of fine bubbles uniformly distributed along the column cross-section of the fluidizer. The antenna type air sparger has previously been used by Meikap et al. (2000). There it has mentioned the generation of well distributed fine bubbles with very less pressure drop values by this design compared to the conventional air distributor (Jena et al., 2009).

Use of various other types of distributor is seen in literature viz. a packed bed of solids of size 2 to 6 mm, distributor containing single nozzle and multiple nozzles , vertical

pipes evenly spaced across a grid containing fine holes, a ring containing large number of holes. The packed type distributor generates uniform mixture of gas and liquid which enters the test section of the three-phase system, results in large pressure drop in comparison to distributor containing different types of air sparger. (Jena et al., 2009)

The bubbles generated by single and multiple nozzle air sparger are generally large in size. The ring type air sparger has unequal distribution of the gas in the liquid

which enters the test section. As the gas-liquid mixture moves up the bed a vigorous contact between the solid phase and liquid and gas takes place in the three-phase system. In the present study, it has been observed visually that the distributor arrangement used allows a uniform flow of gas and liquid to the test section containing fine gas bubbles. In the gas-liquid distributor section of the fluidized bed, the gas and the liquid streams are passed through the perforated grid. The mixing section and the grid ensures that the gas and the liquid are uniformly and equally distributed to the bed.

The gas-liquid disengagement section at the top of the fluidizer is a cylindrical section of internal diameter of 0.26 m and of height 0.34 m , connected to the test section , which allows gas to escape from the system and liquid to be circulated back to the bottom of this section .Fig. 6. For the measurement of pressure drop in the bed, the pressure ports have been provided and fitted to the manometers filled with carbon tetrachloride as the manometric fluids for the measurement of pressure at different ranges. Pressure ports are available at different levels one at bottom and one at the top of the test section. With this arrangement, the wall effect, expanded bed height, distribution of particle concentration and the gas holdup can be studied clearly (Jena et. al, 2009). In actual practice, oil free compressed air from a centrifugal compressor (3 phase, 1 Hp, 1440 rpm) with a receiver and an air accumulator / constant pressure tank used to supply the air at nearly constant pressure gradient as fluidizing gas. The air was injected into the column through the air sparger at a desired flow rate using calibrated rotameter.

Water was pumped to the fluidizer at a desired flow rate using water rotameter.

A centrifugal pumps of capacity (Texmo, single phase, 1 HP, 2900 rpm, discharge capacity of 150 lpm) were used to deliver water to the fluidizer. Three calibrated rotameters with different ranges each for water as well as for air were used for the accurate record of the flow rates. Fig 8 shows the photographic representation of water rotameter and Fig. 9 shows the air rotameter. Water rotameters used were of the range 0to 20 lpm, 5 to 100 lpm and 20 to 200 lpm. Air rotameters were of the range 0 to 10 lpm,0 to 50 lpm and 10 to 100 lpm.

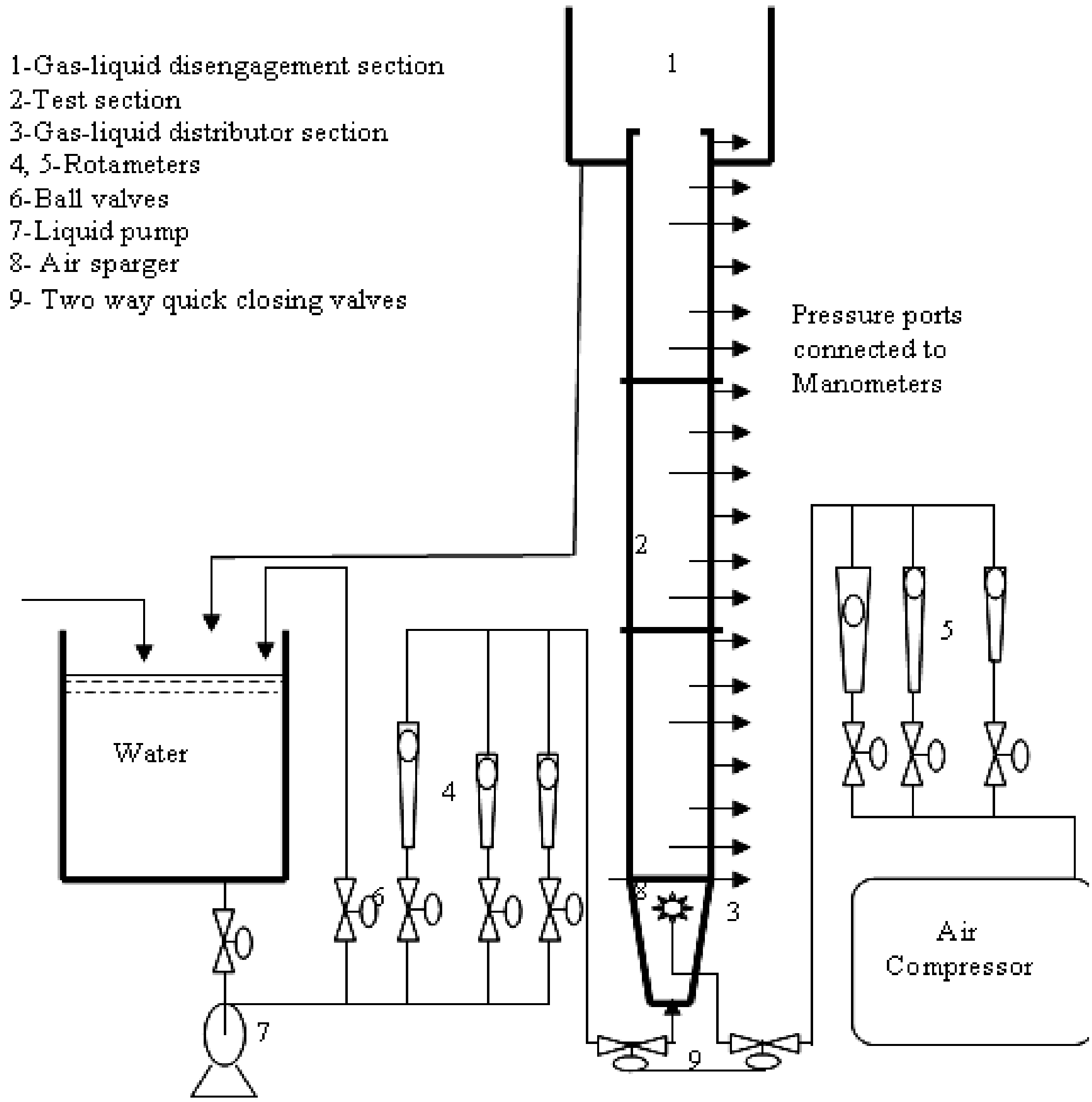

Fig. 1. Schematic representation of the experimental setup of three-phase fluidized bed

Fig. 2 photographic representation of the setup

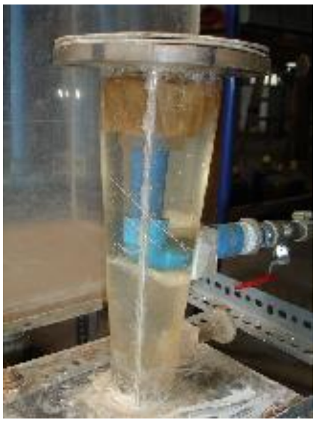

Fig.3 Photographic view of the gas-liquid distributor section

Fig. 4 Photographic view of the distributor plate

Fig.5 Photographic view of air sparger

Fig. 6 Photographic view of disengagement section

Fig. 7 Photographic view of Pump

Fig. 8 Photographic view of water rotameter

Fig. 9 Photographic view of air rotameter

3.1 Experimental procedure

The experiment was carried put using glass-beads as solid particle, water as liquid and compressed air as gas. Initially the column is filled with water and for different gas flow rate variation of pressure drop and bed height is taken for different flow rate of liquid after 5 mins of steady state was attained. The scope of the experiment has been presented in Table 1.
Pressure drop and the expanded bed height was noted. Experiments were conducted at normal temperature of (30±5) ℃. The temperature of the water was also at normal condition ie. At (30±5)° C. The procedure was repeated for different liquid flow rate, particles of different sizes and varying initial static bed heights.

3.2 Scope of Experiment

Table 1.Properties of Bed Material

Sl. No.	Material	d_p, mm	ρ_p, kg.m^{-3}
01	Raschig ring-1	5.9	1670
02	Raschig ring-2	3.3	1670

Table 2.Properties of Fluidizing Medium

Fluidizing Medium	ρ (kg/m^3)	μ (Pa.s)
Air at 30^{o}C	1.166	.00001794
Water at 30^{o}C	995.7	0.000798

Table 3. Properties of Manometric Fluid

Manometric Fluid	ρ (kg/m^3)	μ (Pa.s)
Carbon tetrachloride	1600	0.000942

Table 4: Operating Conditions

Superficial gas velocity	0-0.127 m/s
Superficial liquid velocity	0-0.165 m/s
Static bed heights	15.4 cm , 15.6 cm , 21.4 cm , 26.4 cm , 31.4 cm

CHAPTER- 4

RESULTS AND DISCUSSION

In the present study we examine the hydrodynamic properties ie. pressure drop, minimum liquid fluidization velocity, bed expansion ratio. Raschig rings have been used as a solid phase because of its moderate density and surface to volume ratio o its structure. Experiments were conducted with the gas and liquid flow rates ranging from 0 to 0.127 m/s and from 0 to 0.169 m/s, respectively. Figs. 1 to 6 shows the variation of pressure drop with superficial liquid velocity for gas-liquid–solid system at various bed heights and particle size. Figs. 7 to 13 shows the variation of bed expansion with superficial liquid velocity for gas-liquid-solid system at various gas velocity and static bed height.

4.1. Pressure Drop :

In this study pressure drop was measured from manometer having carbon tetrachloride as manometric fluid in it which are connected to ports at the top and at the bottom of the column. The column completely filled with solid particles up to a desired height and then occupied with water with the initial level of manometer adjusted to have zero. For gas-liquid-solid experiment with little flow of liquid close to zero, the air was introduced slowly and then increased gradually to the desired flow rate after which the liquid flow rate was increased and the readings were noted down.

Figure 10 shows the variation of pressure drop with superficial liquid velocity for gas velocity 0,0.025, 0.0509, 0.076, 0.102, 0.127 m/s of static height 21.4 cm for the particles of size 5.9 mm . The minimum liquid fluidization velocity can be obtained directly from the pressure drop profile vs superficial liquid velocity. It has been found that for different gas velocities as the superficial liquid velocity is increased the pressure drop increases but as soon as it reaches the minimum liquid fluidization velocity then the pressure drop becomes constant. It can be clearly seen from the graph that the minimum liquid fluidization velocity decreases with the increase in gas velocity. This decrease in minimum liquid fluidization velocity may be due to the upward buoyancy force given to the total drag on the particles by the upward movement of gas and liquid. This upward movement of the gas and liquid decreases the total weight or the density of the whole bed due to which the pressure drop decreases. The net velocity or the relative velocity provided gives lower fluidization.

Figure 11 shows the variation of pressure drop with superficial liquid velocity for gas velocity 0 ,0.1019 m/s for static height 31.4 cm. In this figure there is a noticeable difference between the minimum liquid fluidization velocity of the gases. Figure 12 and 13 shows the variation of pressure drop with superficial liquid velocity for gas velocity 0,0.025,0.0509,0.076,0.102 m/s of static height 15.6, 21.4 cm for the particles of size 3.3 mm.

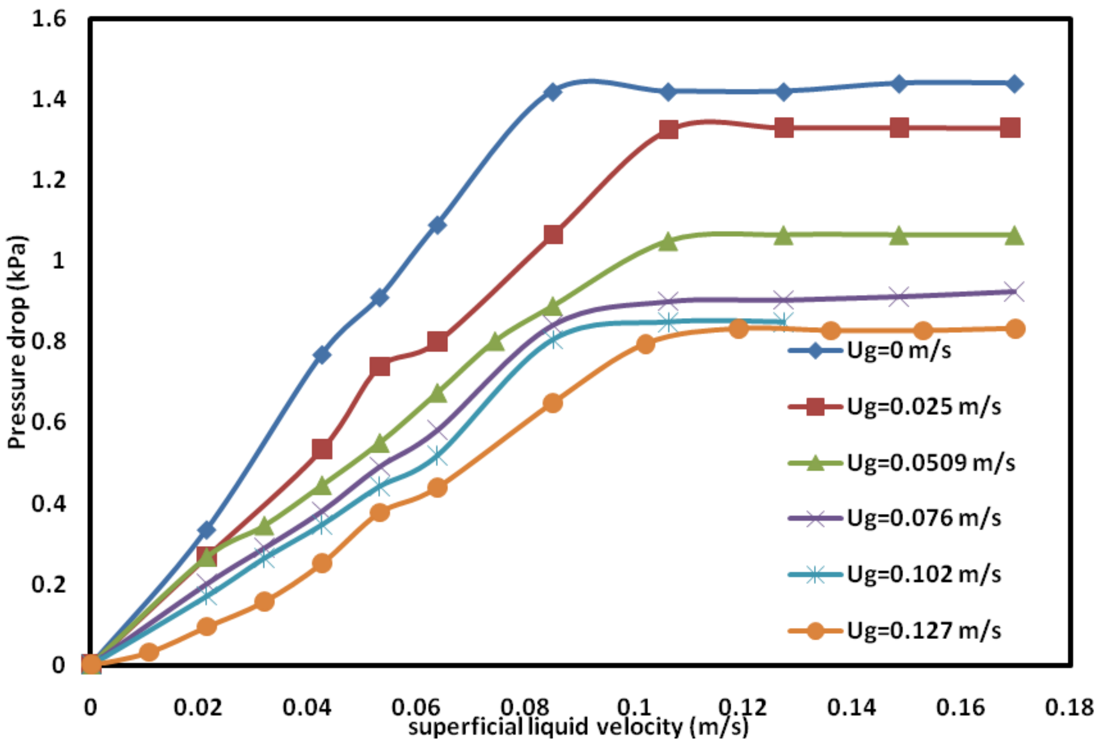

Fig 10 Variation of pressure drop with liquid velocity for different gas velocity for *H*s =21.4cm for dp=5.9 mm

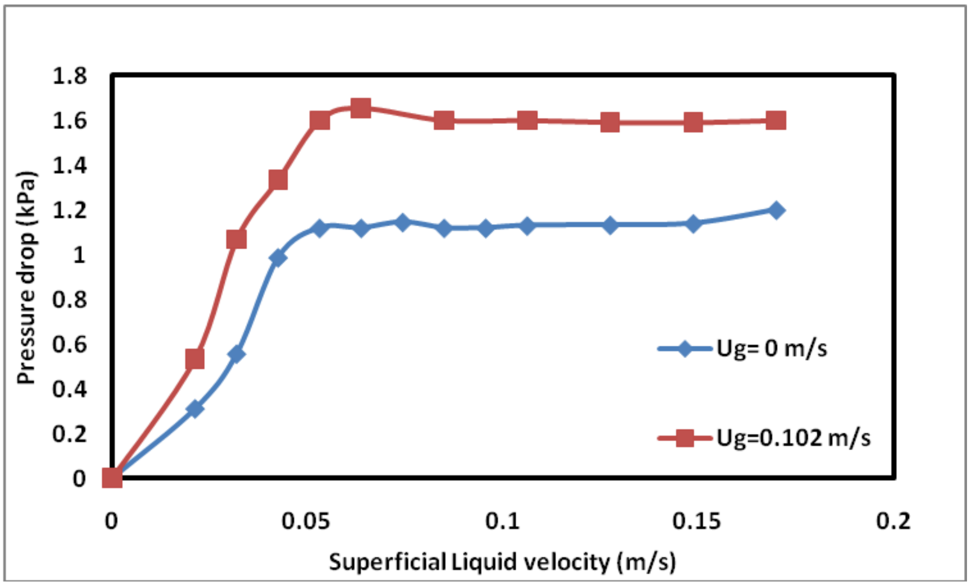

Fig.11. Variation of pressure drop with liquid velocity for different gas velocity at Hs=31.4 cm for 5.9 mm raschig ring.

Fig 12 and Fig. 13 shows the variation of pressure drop with superficial liquid velocity for particle size 3.3 mm. As compared to larger particle size, smaller particle size attains fluidization easily because the upward buoyant force given to the total drag on the particles by the upward movement of gas and liquid is less as compared to bigger particles.

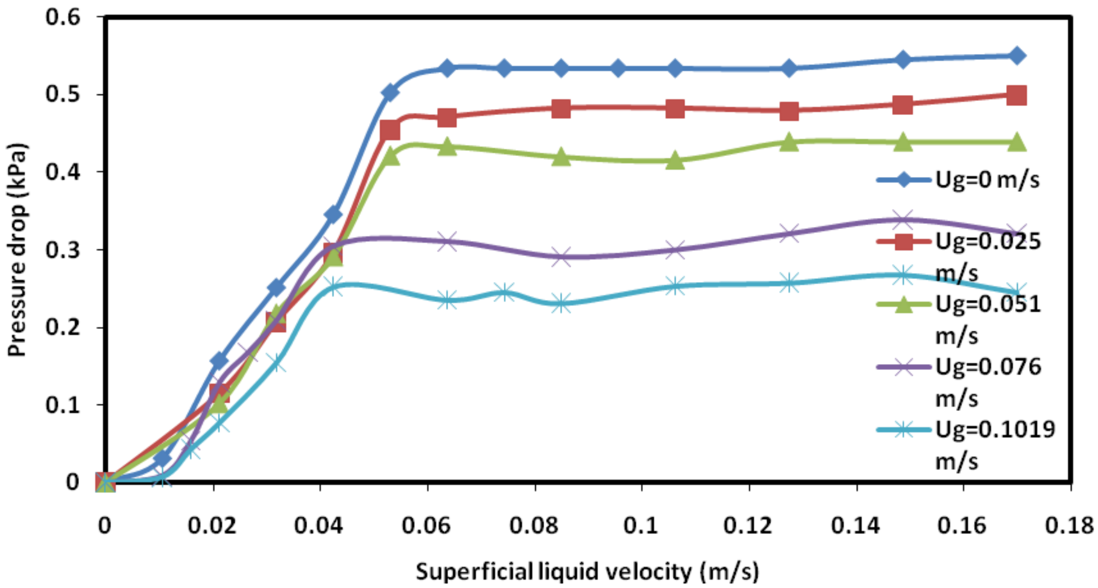

Fig.12. Variation of pressure drop with liquid velocity for different gas velocity at Hs=15.6 cm for 3.3 mm raschig ring

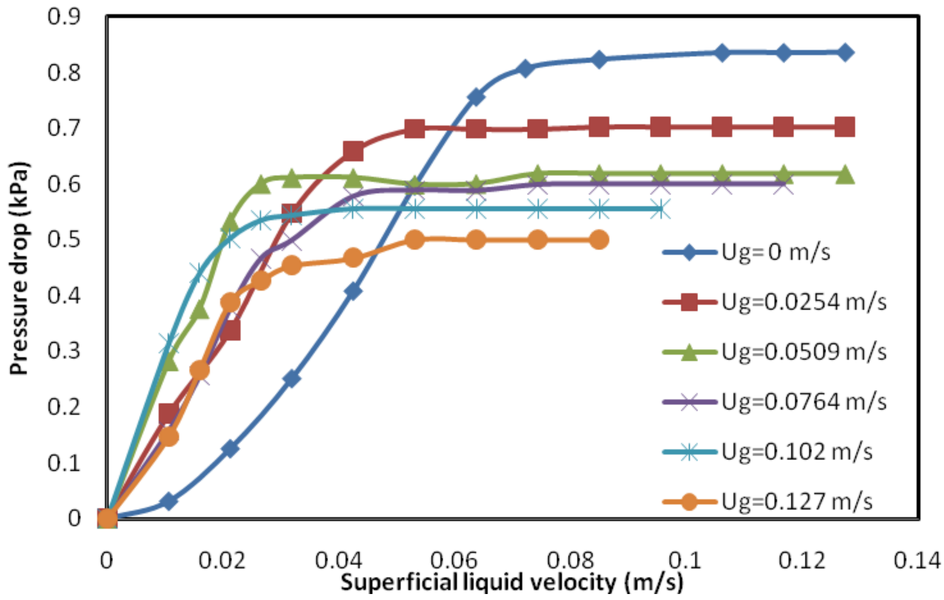

Fig.13. Variation of pressure drop with liquid velocity for different gas velocity at Hs=21.4 cm for 3.3 mm raschig ring.

Figure 14 shows the Variation of pressure drop with liquid velocity for different bed height at *U*g =0.1019 m/s for 5.9 mm raschig ring. The minimum liquid fluidization velocity can be seen from the graph. It has been found that bed mass has no effect on minimum liquid fluidization velocity. The minimum liquid fluidization velocity has been found to be 0.053 m/s for different static bed height. Similarly for figure 15 shows the Variation of pressure drop with liquid velocity for different bed height at *U*g =0.076 m/s for 3.3 mm raschig ring. Here also the minimum liquid fluidization velocity is not getting affected by bed mass. The minimum liquid fluidization velocity is found to be to 0.0315 m/s for different static bed height and for figure 16 it is found to be 0.0265 m/s.

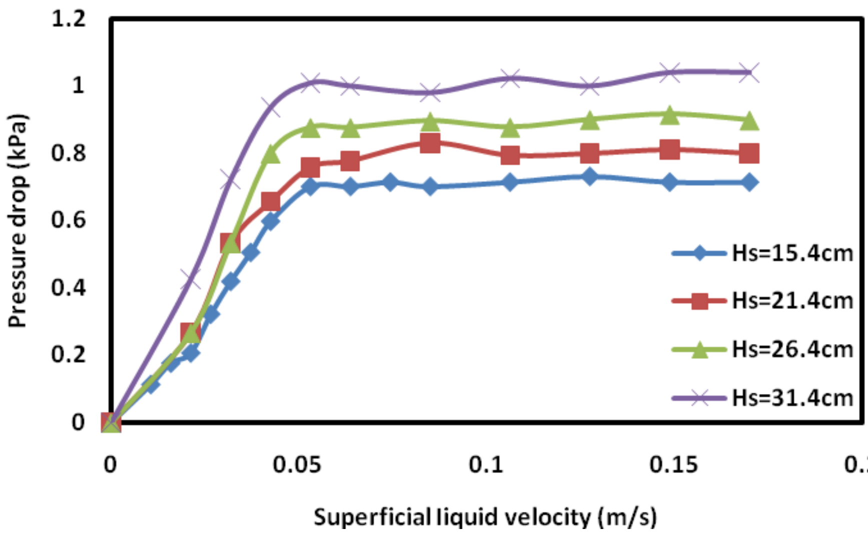

Fig.14. Variation of pressure drop with liquid velocity for different bed height at *U*g =0.1019 m/s for 5.9 mm raschig ring.

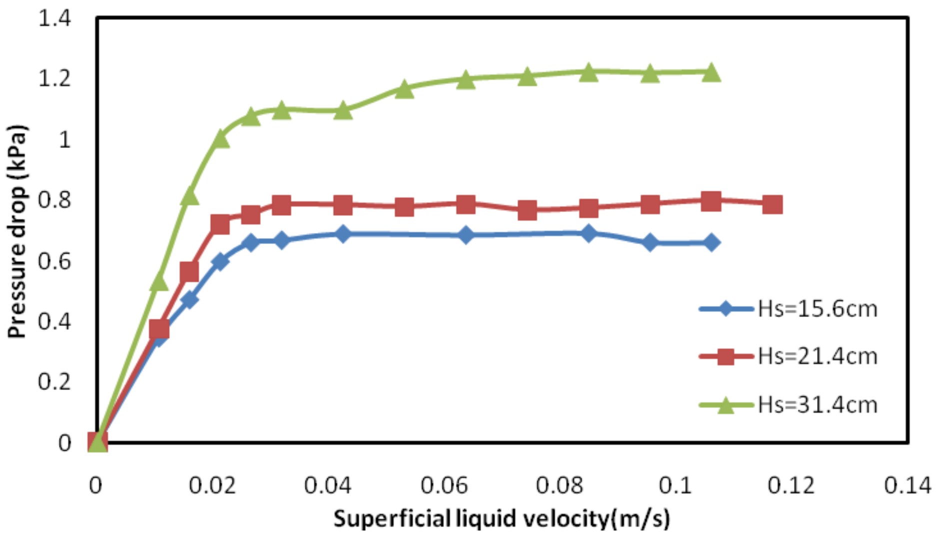

Fig.15. Variation of pressure drop with liquid velocity for different bed height at *U*g =0.076 m/s for 3.3 mm raschig ring.

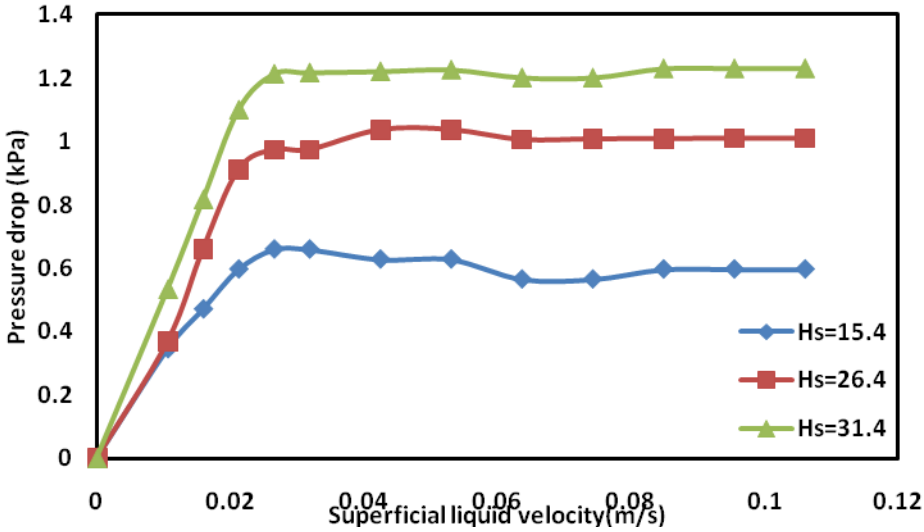

Fig.16. Variation of pressure drop with liquid velocity for different bed height at Ug = 0.063 m/s for 3.3 mm raschig ring.

Fig. 17 shows the variation of pressure drop with liquid velocity for different particle size .For particle size of 3.3 mm it attains fluidization at 0.064 m/s while the particle size of 5.9 mm it attains fluidization at 0.0849 m/s .Thus we found that the larger size particles require much pressure force for fluidization. Hence minimum liquid fluidization increases with the increase in particle size

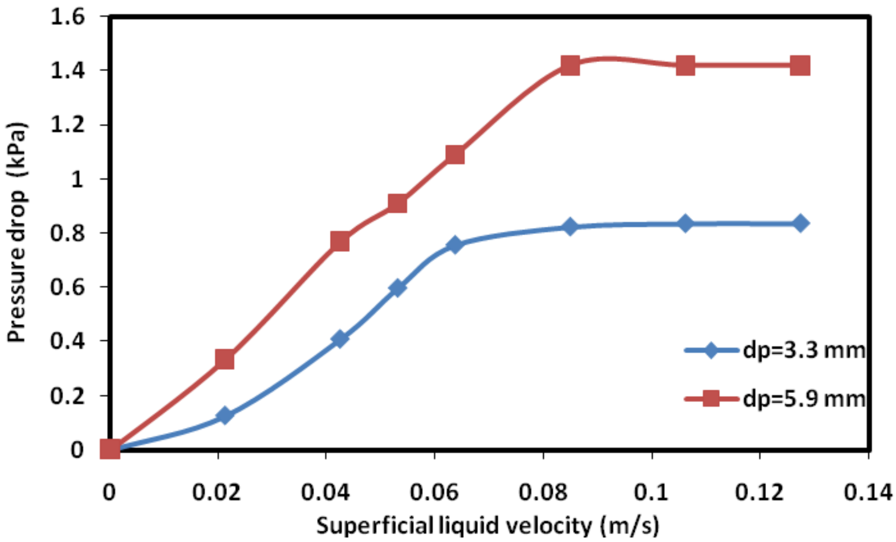

Fig 17. Variation of pressure drop with liquid velocity for different particle size at Vg =0 m/s for Hs =21.4 cm

4.2 Minimum liquid fluidization velocity

Minimum liquid fluidization velocity is that velocity at which the bed reaches its fluidizing condition. In the present work it has been found by visual observation.

Fig. 18 shows the variation of minimum liquid fluidization velocity with gas velocity for particle size 5.9 mm and for static height 21.4 cm. The minimum liquid fluidization velocity decreases with the increase in gas velocity. This decrease in minimum liquid fluidization velocity is due to the bubble supported fluidization which generally decreases with the increase in gas velocity.

Fig.19 shows the the variation of minimum liquid fluidization velocity with gas velocity for particle size 3.3 mm and for static height 21.4 cm. Here also the minimum liquid fluidization velocity decreases with the increase in gas velocity. The decrease in minimum liquid fluidization velocity is due to the bubble supported fluidization which generally decreases with the increase in gas velocity.

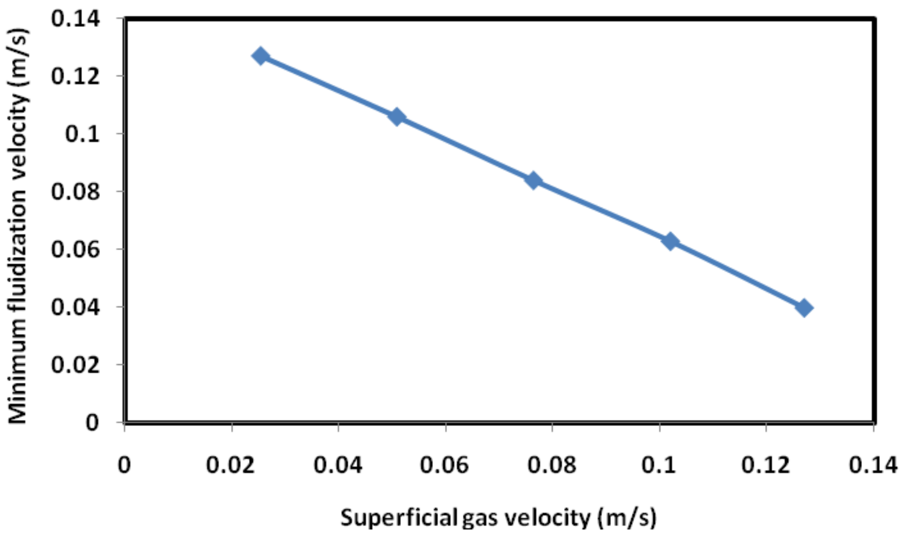

Fig.18. Variation of minimum liquid fluidization velocity with gas velocity

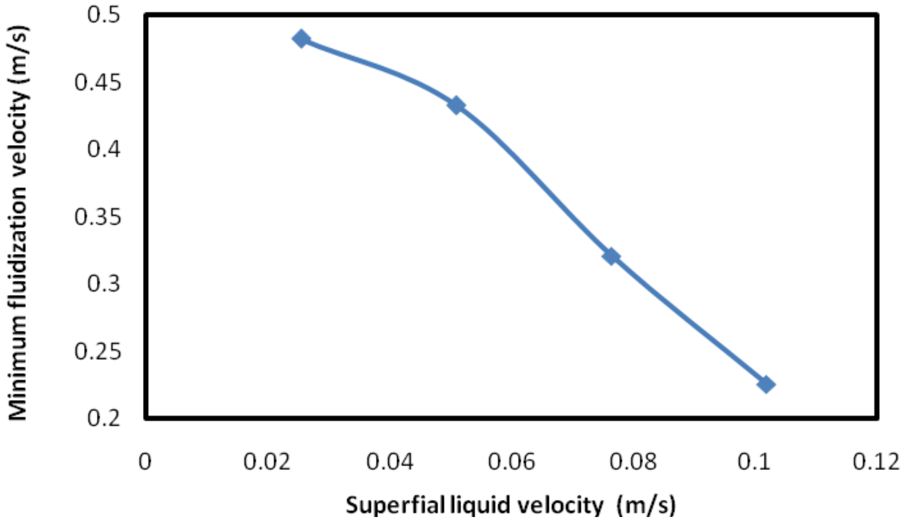

Fig.19. Variation of minimum liquid fluidization velocity with gas velocity

Fig. 20 shows the Variation of minimum liquid fluidization velocity with gas velocity for different particle sizes at a constant static bed height of 21.4 cm. From the graph we can see that as the gas velocity is increased minimum liquid fluidization velocity decreases. From the graph we can conclude that the minimum liquid fluidization velocity increases with the increase in particle size.

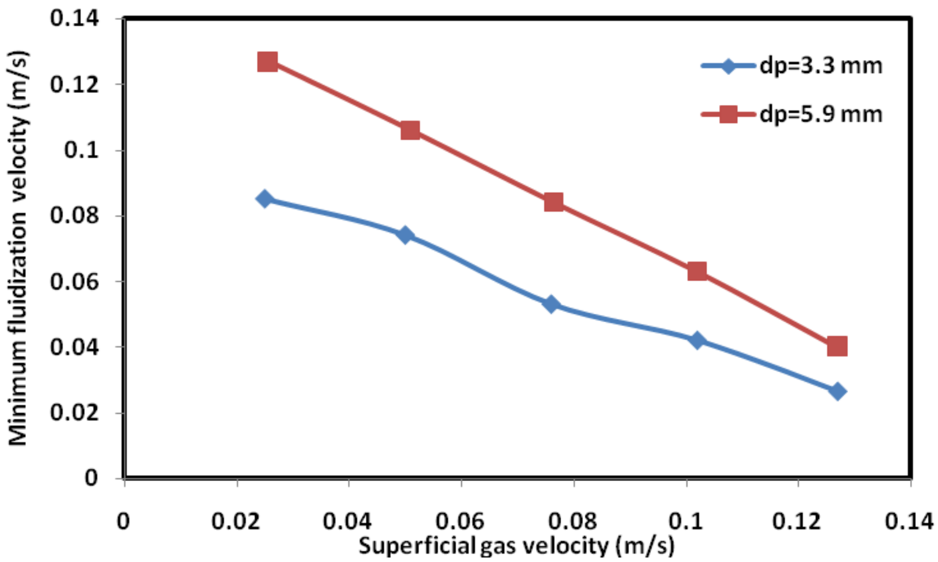

Fig.20. Variation of minimum liquid fluidization velocity with gas velocity

4.3 Bed expansion

Bed expansion ratio is defined as the ratio of the fluidized bed height to the static bed height. In the present work it has been found by visual observation. The bed expansion ratio was studied by varying the liquid velocity at constant gas velocity.

Fig. 21 shows the variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 21.4 cm for particle size 5.9 mm. It has been found that the bed expansion ratio increases with the increase in gas and liquid velocity. Initially for lower gas velocities the bed expansion ratio is greater but as the gas velocity is increased the expansion ratio also increases i.e. it rises above the column so there may be chances of particle loss. So for higher gas velocity it is not fluidized for higher liquid velocity. Fig. 22 shows the variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 31.4 cm for DP=5.9 mm. Higher bed expansion ratio is achieved faster for solids operated at higher gas velocity. This is due to the increased gas holdup in the column which accounts for further bed expansion.

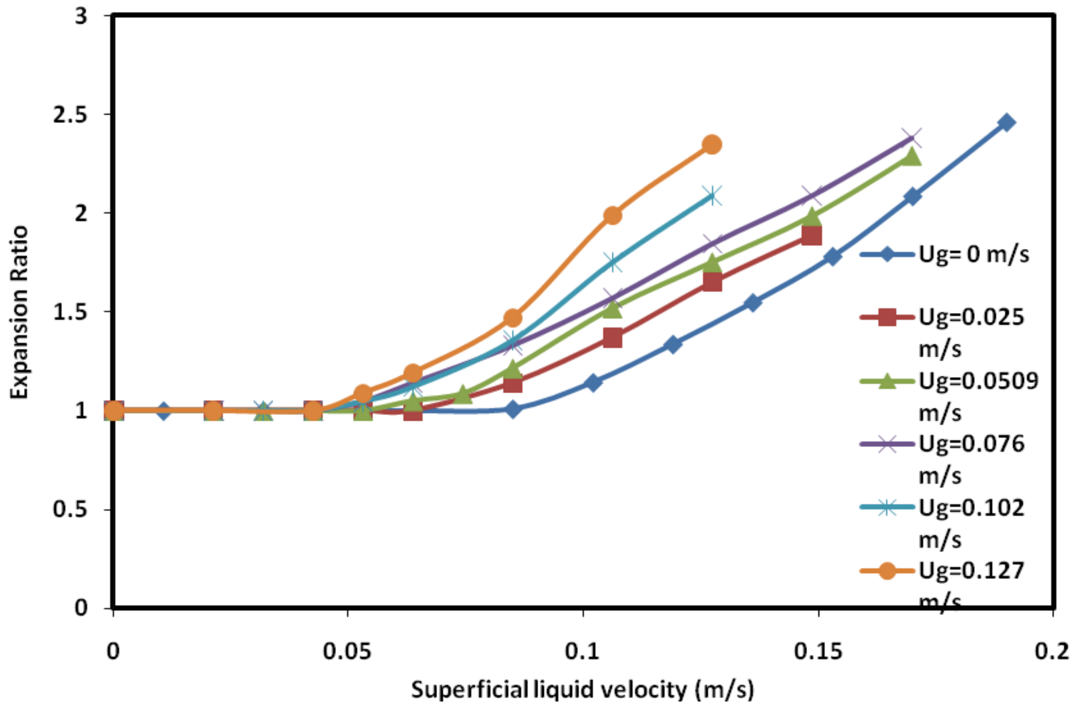

Fig. 21. Variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 21.4 cm for d_p= 5.9 mm.

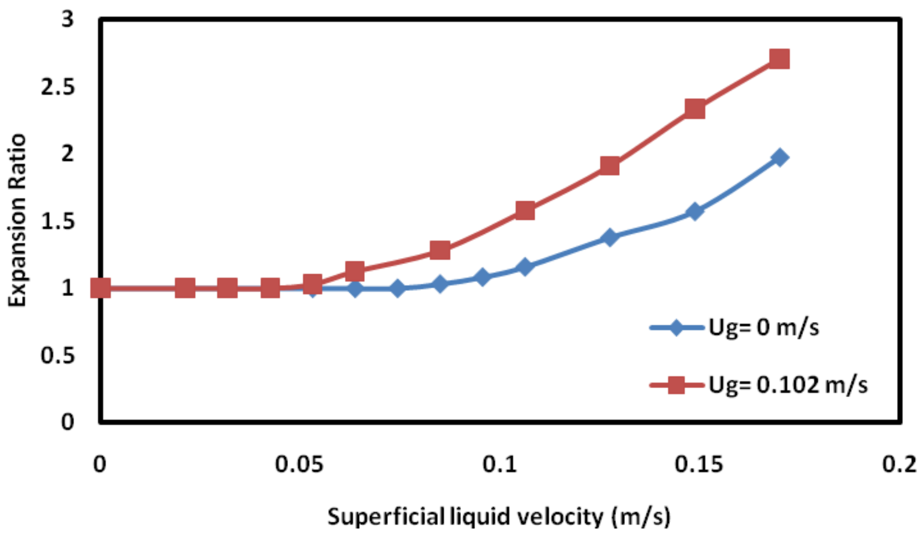

Fig.22. Variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 31.4 cm for d_p=5.9 mm.

Fig. 23 and 24 shows the variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 15.6 cm, 21.4 cm for particle size 3.3 mm. It has been observed that for smaller sized particles bed expansion ratio is slightly higher than the larger sized particle due to its less weight.

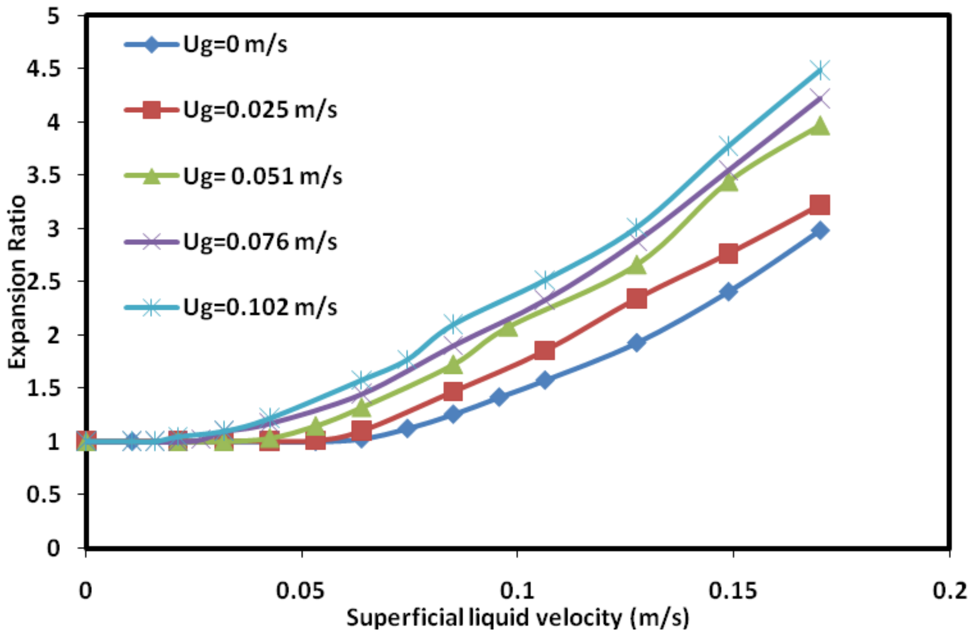

Fig. 23. Variation of bed expansion ratio with liquid velocity for different gas velocities at *H*s = 15.6 cm for d_p=3.3 mm

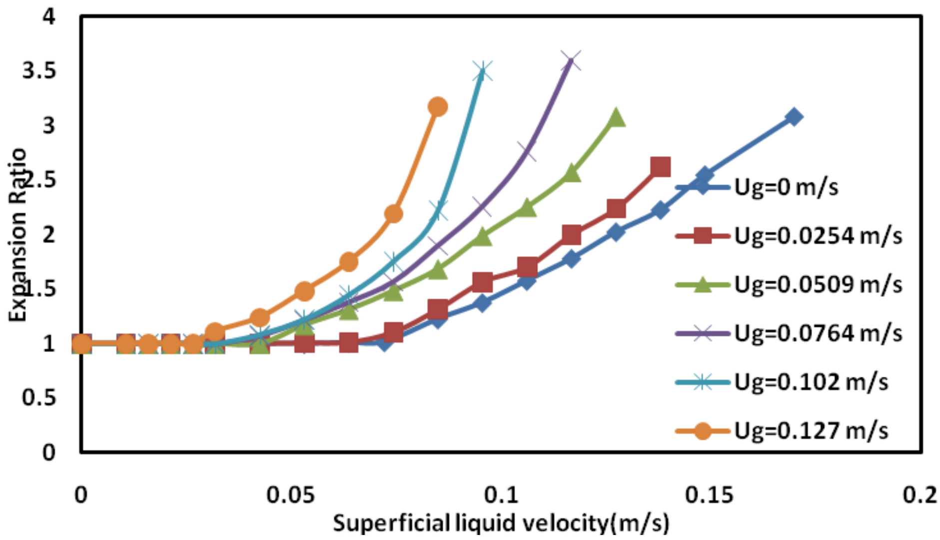

Fig. 24. Variation of bed expansion ratio with liquid velocity for

Different gas velocities at *H*s = 21.4 cm for DP =3.3 mm

The following figure 25, 26, 27 shows the variation of bed expansion ratio with liquid velocity for different static bed height for constant value of gas velocity and for particle size of 5.9 mm and 3.3 mm. It has been observed that the expansion ratio is independent of initial static bed height.

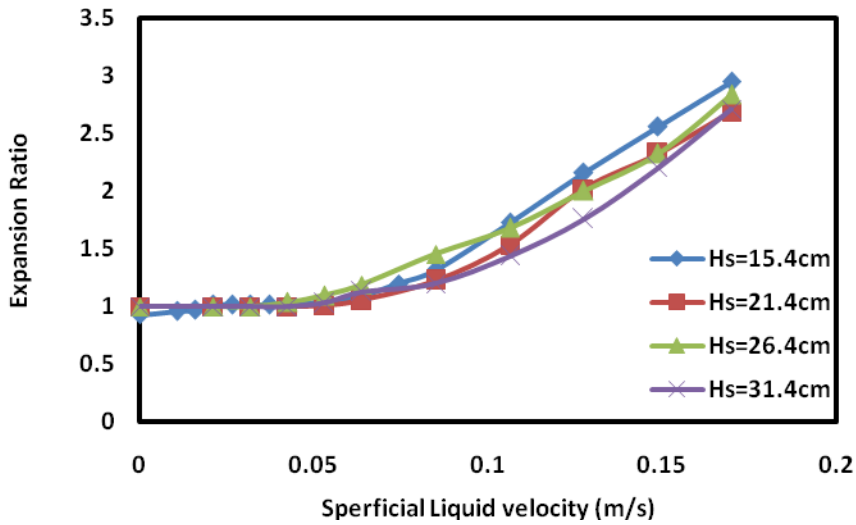

Fig.25. Variation of bed expansion ratio with liquid velocity

for different Hs when Ug=0.1019 m/s for d_p= 5.9 mm

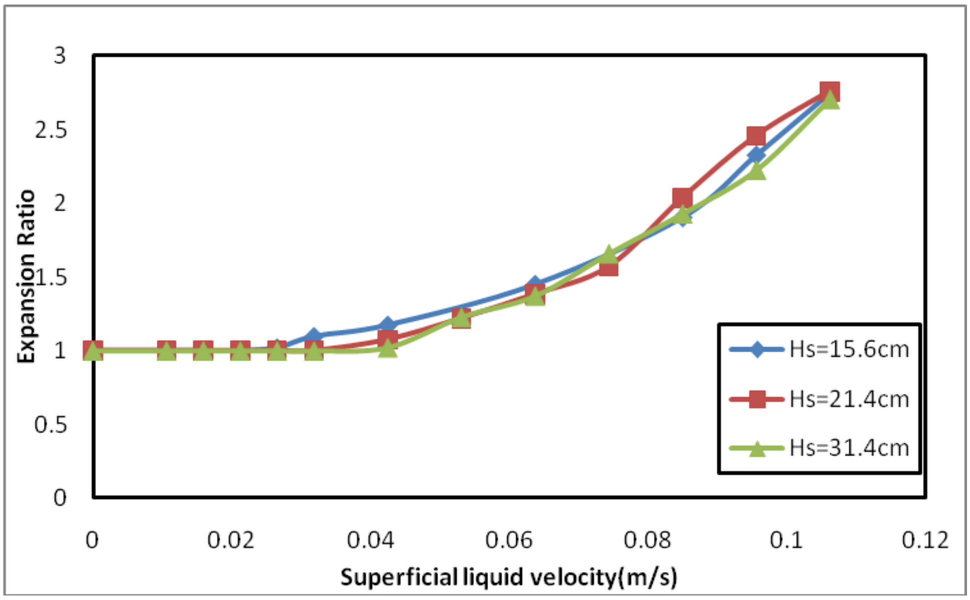

Fig. 26. Variation of bed expansion ratio with liquid velocity for different Hs when Ug=0.076 m/s for d_p=3.3 mm

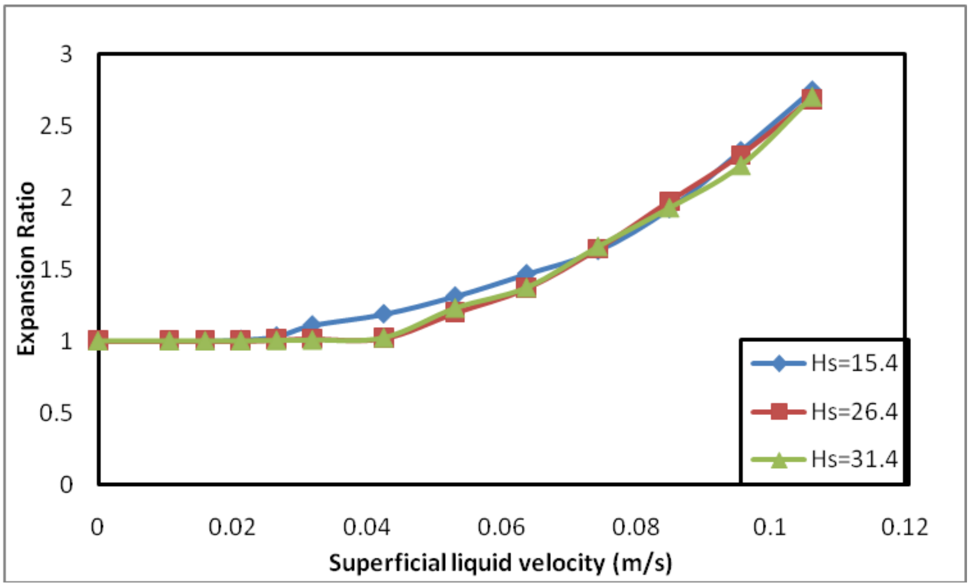

Fig. 27. Variation of bed expansion ratio with liquid velocity for

different Hs when Ug=0.637 m/s for d_p= 3.3 mm

Fig. 28 shows the variation of bed expansion ratio with liquid velocity for different for particle size of 3.3 mm and 5.9 mm at zero gas velocity. Bed expansion ratio decreases with the increase in particle size.

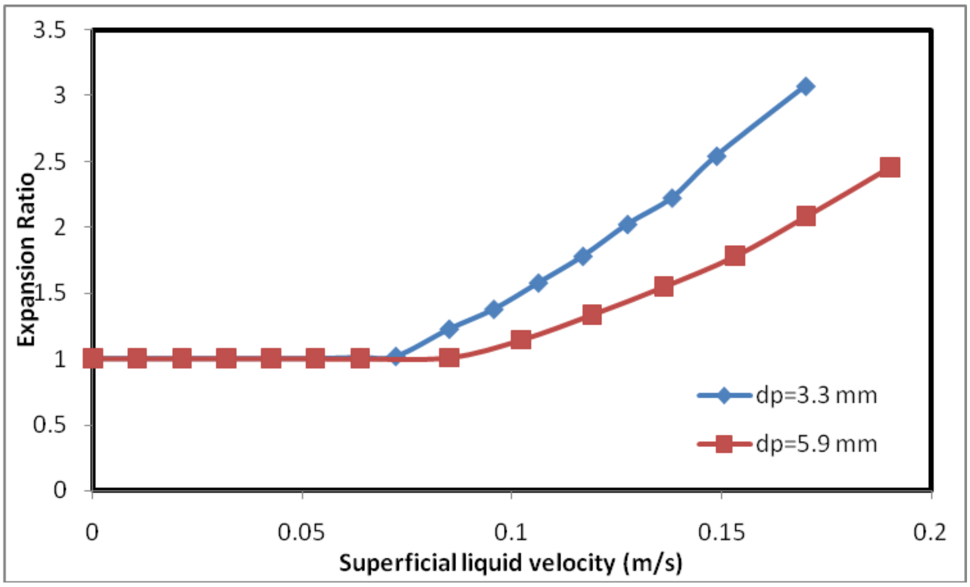

Fig. 28. Variation of bed expansion ratio with liquid velocity for different for particle size of 3.3 mm and 5.9 mm at Ug=0 m/s

CHAPTER-5

CONCLUSION

The hydrodynamic study of the three- phase fluidized bed has been determined and following conclusions are drawn.

5.1 Pressure Drop

- The bed pressure drop decreases with the increase in gas velocity.
- With the increase in bed mass the pressure drop increases.
- With the increase in particle size the pressure drop increases.

5.2 Minimum liquid fluidization velocity

- The minimum liquid fluidization velocity decreases with the increase in gas velocity.
- Minimum liquid fluidization velocity increases with the increase in particle size.
- Minimum liquid fluidization velocity is independent of initial static bed heights.

5.3 Bed expansion ratio

- Bed expansion ratio increases with the increase in gas and liquid velocity.
- Bed expansion ratio is independent of initial static bed height.

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